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Summary

This paper outlines some of the design approaches being employed during a current DOE/NASA supported study directed at developing an improved air separation process for the production of oxygen enriched air for MHD combustion. The ultimate objective of the study is to arrive at conceptual designs of air separation plants, optimized for minimum specific power consumption and capital investment costs, for integration with MHD combined cycle power plants. Final results will be published following completion of the contract in mid-1980.

Introduction

The benefits of oxygen enrichment at low oxidizer preheat temperatures on combined cycle MHD power plant performance were discussed in references 1 and 2. Medium purity oxygen plants with low power consumption (208 kW-hr/ton of equivalent pure oxygen) were identified.[2] The study was based on technology [3,4,5] used in a 60% purity oxygen plant, built by Linde AG for the Thyssen Steel Works, Federal Republic of Germany. This plant has been operating reliably since 1972, delivering a 1.05 bar enriched air stream of 170,000 scfm with 30% oxygen content at a specific power consumption of 224 kWh per ton of equivalent pure oxygen.

Use of oxygen enrichment in combined cycle MHD plants eliminates the need for high temperature air heaters which are faced with severe corrosion problems. As a result, the development time for commercializing combined cycle MHD plants could be potentially shortened.

For the production of the large quantities of oxygen required by the MHD combined cycle plants, up to 6000 tons of oxygen/day for the baseload MHD combustor ozidizer stream, the cryogenic air separation process is the most economical and it will remain so for the foreseeable future. For this reason the study concentrates on further improvements in cryogenic air separation plants by means of parametric tradeoff studies between the various process variables.

Basic Air Separation Plant

Air separation plants use rectification columns consisting of an upper, low pressure compartment (low pressure column) and a lower compartment operated at higher pressure (medium pressure column). Such plants are referred to as "double column" plants. Figure 1 shows a schematic of a conventional double column plant. Its main components are:

-Air compressor, (CP)
-Adsorbers for the removal of water and carbon dioxide, (WS, MS1, MS2)
-Expansion turbine for the production of refrigeration, (T)
-Heat exchangers for cooling the air to liquefaction temperature and warming the oxygen and nitrogen products to ambient temperature, (E)
-Rectification columns for the separation of air into O_2 and N_2 on the basis of vapor-to-liquid heat exchange and mass transfer, (C1, C2)

In many plants the removal of water and carbon dioxide is combined with the cooling of the air. This is done in reversing heat exchangers or regenerators. The flow paths of the air and product streams can be traced on Figure 1. Filtered atmospheric air is compressed in compressor, CP, to a pressure which is determined by the requirements of the particular process, usually 4 to 6 bars. It is then passed through the water separator, WS, and adsorbers, MS, to the heat exchangers, E, and cooled to about 100 K with the aid of expansion turbine, T. The cold air is injected as a saturated vapor into the medium-pressure column, C1, where it is separated into a nitrogen fraction and the so-called rich liquid fraction, which contains about 40% oxygen. The separation is obtained by contacting the rising vapor with liquid nitrogen flowing down from the condensor-evaporator, K. This causes the liquid collecting in the kettle at the bottom of C1 to become enriched in oxygen while the vapor, which is condensed at the top of C1, is enriched in nitrogen. Both fractions leave C1 in the liquid state and flow through expansion valves to the low-pressure column, C2. In C2 vapor formed in the condensor-evaporator, K, rises while liquid nitrogen flows down from the top of the column. The liquid flowing down to the bottom of C2 is enriched in oxygen and the vapor rising to the top of C2 is enriched in nitrogen. Both fractions leave column C2 in the gaseous state, then are heated in heat exchanger, E, and flow to the consumer. Note that for MHD power plant applications an atmospheric product stream will have to be compressed to combustor operating pressure. Because of the increased oxygen concentration in this stream, specially designed compressors will be required for this purpose.

The physics of the separation process is based on the fact that the vapor pressure of nitrogen is higher than that of oxygen at the same temperature. Therefore, vapor which is in contact with liquid is richer in nitrogen than the liquid. To bring about the heat and mass transfer between the vapor and

liquid fractions required by the air separation process columns C1 and C2 are equipped with sieve trays as shown in Figure 2. The vapor rises through the holes of the sieve trays and bubbles through the liquid flowing over the sieve trays. For low purity product requirements, up to 40% oxygen, the top column sieve trays can be omitted.

Fundamental Operating Theory

The ratio of nitrogen contained in the vapor to the nitrogen contained in the liquid of a nitrogen-oxygen system in equilibrium decreases with rising pressure and increases with falling pressure. Due to this fact the separation of air into oxygen and nitrogen becomes more difficult at higher pressure and vice versa. This problem can be alleviated by reducing the product purity. In the condenser-evaporator, K, oxygen is evaporated and nitrogen condensed as already mentioned. Due to the difference in vapor pressures, a higher pressure is required for the condensation of nitrogen than for the evaporation of oxygen at the same temperature. This pressure difference requirement between the medium and the low-pressure column accounts for the major portion of total energy consumption by the cryogenic air separation process.

Figure 3 shows the pressure ratio between the lower and upper columns, including temperature difference at the condensor-evaporator and pressure drop in the piping, as a function of the product purity, with pressure at the bottom of the low pressure column as a parameter. The pressure ratio increases with increasing product purity and decreasing pressure in the low pressure column, and vice versa. It can be seen that the smallest pressure ratio requirement occurs for a pressure of 8 bar absolute and 30% oxygen purity in the low pressure column, and the largest pressure ratio is required for a pressure of 1.4 bar absolute in the low pressure column and a purity of nearly 100% oxygen.

If one assumes an ideal air separation plant in which there are no friction losses, no heat losses, no air losses, a 100% oxygen recovery and heat exchange with no temperature difference, then the required pressure ratio determines the (theoretical) minimum energy consumption for the process.

The energy consumption of an air separation plant for the production of oxygen with a purity of 99.5% O2 is shown in the following table:

Table I: Energy Consumption Schedule for Pure Oxygen Production

Theoretical energy consumption for the cryogenic process	45.1 %
Air flow losses	3.2 %
Pressure drop losses	18.3 %
Heat exchange losses	1.5%
Insulation losses	0.6 %
Oxygen recovery losses	2.3 %
Losses for inefficiency of the air compressor	29.0 %
Total energy consumption	100.0 %

This table shows that the three essential energy consumers for the separation of air are the theoretical consumption, the pressure drop losses and the inefficiency of the air compressor.

Process Variations

We have already shown (Figure 3) that the column pressure ratio, and hence the theoretical consumption, can be reduced by decreasing the product purity, 99.5% in Table 1, and by increasing the absolute pressure of the process. Both conditions are based on the fact that the already described double-column process is used.

There is no other cryogenic air separation process which results in a lower theoretical energy consumption, because the reflux nitrogen must always be condensed by evaporating oxygen. It is a general requirement of the cryogenic air separation process that reflux liquid and rising vapor must be taken from the medium to be separated. There are some sophisticated ways for decreasing the theoretical energy consumption. For example, one may introduce an interconnecting cycle, which is operated with a thermodynamically advantageous medium, e.g. argon. Another idea is to combine the separation and condensation in a special unit in which condensation, evaporation and separation takes place. The problems with this unit are limited recovery of products and high investment costs. All such processes are complicated and expensive. Therefore, the study is limited to variations of the basic double column process described above.

The pressure drop losses referred to in Table 1 can be reduced by enlarging of pipe diameters and by increasing the working pressure. The higher working pressure also permits a higher gas throughput rate at limited pressure drop. Based on this, better heat transfer, which results in smaller heat exchangers, can be obtained. These advantages would be reduced to some extent by the greater difficulty in separating the air components at high pressure operation.

The isothermal efficiency values of actual air compressors are shown in Figure 4 for a range of pressure ratios. Data trends indicate that the isothermal efficiency decreases with increasing pressure ratio.

In order to reduce power consumption of an air separation plant the total losses can be minimized by combining components in an optimal way. The individual components can be optimized, e.g. the power consumption of an air compressor can be reduced by use of high efficiency axial stages, variable speed drives, use of variable stator guide vanes, and increasing the number of compressor stages with intercoolers. This has been the approach taken in our study, which included analysis of a large number of double-column processes.

The energy consumption of these processes can be compared by use of certain parameters as follows:

$$C = \frac{P}{E} \qquad - (1)$$

where
C = specific energy
P = power consumption
E = quantity of enriched air with a specified oxygen content

If formula (1) is applied, an answer to the question of optimum process selection can be obtained. We will not, however, explain why this is the optimum process. It is heretofore not possible to get a complete answer but a closer analysis of

the formula will give some more information. We separate formula (1) into three factors as follows:

$$\frac{P}{E} = \frac{P}{A} \cdot \frac{A}{O} \cdot \frac{O}{E} \qquad - (2)$$

where

P = power consumption $\frac{P}{A}$ = power factor;

A = air flow $\frac{A}{O}$ = air factor

O = product oxygen flow $\frac{O}{E}$ = product factor;

E = enriched air flow $\frac{P}{E}$ = specific energy

A more detailed understanding of the above factors may be obtained from Figure 5. This figure shows an air separation unit (ASU) as a black box. Air is fed in by the compressor, CP, at a certain pressure. Oxygen (O) leaves the black box as the main product, while two nitrogen streams (N1 and N2) are by-products. The power factor is defined on the basis of the formula for isothermal compression. An isothermal efficiency factor can be applied, but this will not be done in this paper, since only relative comparisons are desired. A bonus will be given for the products leaving the black box at increased pressure, as indicated by the negative terms in the power factor expression, since power may be recovered from these streams by expansion turbines. The air factor is the ratio of air flow to product oxygen flow. It represents the inverse of the oxygen recovery.

The product factor is the ratio of product oxygen flow to the flow of enriched air. It is obvious from Figure 5 that the oxygen concentration of the product stream need not be the same as the oxygen concentration of the enriched air. We therefore have the possibility of optimizing the air separation plant independently of the requirements for the MHD-system. If the oxygen concentration of the product stream is higher than that of the enriched air we must add atmospheric air to obtain the desired oxygen concentration.

The quantity of oxygen required in the enriched air stream is determined by the desired O_2 concentration. Since we take a portion of this oxygen quantity from atmospheric air, we can produce a smaller amount of oxygen enriched product with the air separation plant. Moreover, the required quantity of product will decrease as product oxygen concentration increases. By considering the mixing of air and product, it is not difficult to conclude that the ratio of product flow to enriched air flow must be

$$\frac{O}{E} = \frac{(e-0.2095)}{(o-0.2095)} \qquad - (3)$$

where

e = oxygen concentration in the enriched air
o = oxygen concentration in the product stream from the ASU
0.2095 = oxygen concentration in atmospheric air

Figures 6, 7, and 8 show the above described factors, each as a function of the product oxygen purity. The power factor relationship is shown in Figure 6. The plotted power factor values were calculated for several previously designed double-column plants. The various power factors can be connected by two parallel straight lines. The upper line connects power factor values for plants which operate at near atmospheric pressure in the low pressure column. Two power factor values, one at 70% and one at 95% product oxygen purity fall below the line. The value at 70% is for the improved Linde/Lotepro process which uses a separate product evaporator and a reflux condenser. The value at 95% purity represents the Linde/Lotepro two column process discussed in reference 2. The lower line in Figure 6 approximately correlates power factor values for plants with low pressure column pressures ranging from 5 to 9 bars. Since only relative comparisons between the various processes were desired, the power factors were computed on the basis of isothermal compression. Absolute power factor values can be obtained by multiplying the relative values by the isothermal compressor efficiency.

Figure 7 shows the line correlating calculated air factor values for several double column plants.

The product factor is shown in Figure 8. The two curves were calculated by using formula (3). The lower curve shows that the product factor for an enriched air purity of 30.6% decreases from unity at a product purity of 30.6% oxygen to 0.1223 at 99.5% product purity. The upper curve shows the product factor for an enriched air purity of 41%.

Figure 9 shows the specific energy, P/E, which is the product of the three factors. There are two curves. The upper one connects all double-column plants with a pressure in the low-pressure column, near atmospheric. The curve has a minimum at about 70% oxygen purity. This result is surprising. It is based on the strong decrease of the product factor with the increasing power factor. Or, in other words, the quantity of air to be processed in an air separation plant decreases up to 70% oxygen purity at a faster rate than the power expenditures increases. The absolute specific energy value of the 70% point corresponds to approximately 195 kW-hr per short ton of equivalent pure oxygen.

The lower curve connects specific energy values for theoretical double column plants with a pressure in the low-pressure column between 5 and 9 bars. This curve has a minimum at 60% oxygen purity. It is important to remember that these results were obtained by assuming perfect isothermal compression of the inlet air stream and perfect power recovery from the pressurized nitrogen by-product stream. Actual turbomachinery efficiencies, when they become available for each case, are expected to raise the specific energy values for medium pressure plants to or even above those for the best atmospheric plants. The points marked AN and AD in Figure 9 will be discussed in connection with Figure 12.

Capital Costs

Figure 10 shows the capital cost of air separation plants as a function of the oxygen purity for two levels of product pressure. These costs are based on relative figures starting with a cost ratio of 1 for a normal plant which produces 99.5% of pure oxygen. The influence of the investment costs depends on the depreciation period and the rate of interest. Generally the influence of the investment costs on the product oxygen costs are of minor importance, if a depreciation period of about 20 years is used. Minimum capital investment is required for a process producing low pressure product at 70% oxygen purity. In absolute terms,

the turnkey cost for a baseload scale 70% purity air separation plant would amount to about 15% of the total cost for the combined cycle MHD power plant. Note that capital costs are higher for plants delivering product streams at elevated pressure.

Peaking Operation

An air separation plant can be used as a peak shaving unit in order to reduce power consumption during periods of high electric power demand. Figure 11 shows the relative flow values corresponding to an alternating oxidizer flow demand and the steady product flow delivery from the air separation plant. High demand is shown for weekday operation with low demand at night and on weekends.

In a standard air separation plant the varying oxygen enriched product flow can be obtained by adapting the air separation plant to the demand. This method requires a power expenditure proportional to the product flow. Peak shaving is a way to vary the product flow without proportional power increase. This yields a power saving during daytime operation when unit power costs are high. Cryogenic air separation plants can be adapted for peak shaving by producing and storing excess liquid oxygen at night, evaporating it during the day and adding it to the main product flow.

There is another method of peak shaving which involves compression of the excess oxygen produced during periods of low power demand to a high pressure and storage in pressure vessels. During the day the stored oxygen can be withdrawn from the vessels and passed through an expansion turbine to the MHD plant. This system will be reliable and efficient, but the cost of the pressure vessels and the high pressure oxygen compressors is very high.

There is a third method, described below, which is based on the alternating storage of liquid oxygen and liquid air. This method allows for conservation of the energy contained in stored liquid oxygen, which may range from 0.35 kWh/kg to 0.56 kWh/kg. The air separation plant supplies a steady oxygen flow which is the arithmetic mean of the oxygen demand during a one-week period as is shown in Figure 11. During periods of low demand the excess oxygen production (which is the difference between the mean and actual product flow) is stored as a liquid. It is liquified by evaporation of liquid air which has been stored in a second tank. During the day the difference between the peak product flow and the mean flow is supplied by evaporating oxygen which is taken from the oxygen tank. Gaseous air is condensed by the evaporating oxygen and collected in the liquid air tank.

Figure 12 shows the flow sheet of the air separation plant combined with oxygen tank, TO, liquid air tank, TA, and the product condenser-evaporator, K2. The heat transfer between oxygen and air takes place in the condensor-evaporator, K2. During periods of low power demand this system has a higher power factor and a higher air factor than a normal air separation plant. The converse is also true. For the evaporation of the oxygen during the day an equivalent air stream has to be used and stored in the liquid state which results in an unbalanced flow through the heat exchanger. This causes refrigeration losses which must be covered by an increased expansion turbine output. The additional insulation losses of the tanks must also be provided for.

The specific energy factors for this system have been plotted as shown in Figure 9. The symbol AD shows the specific energy factor for daytime operation and AN shows the specific energy factor for night time operation. These two factors are based on an assumed relative product flow, to the MHD combustor, shown in Figure 11. A fictitious power consumption can be calculated by placing a value on daytime and nighttime power. For example: if the nighttime power has a value of only 0.5 compared to the daytime power, a saving of about 7-9% can be obtained.

The system described above can be improved by varying the mean flow rate. The plant can be operated at reduced mean flow rate during high demand periods and vice versa. Then the differences between demand and supply flow rates will increase, requiring larger tanks. Based on the above assumptions an additional 5% power saving can be obtained.

The design of an alternating oxygen/liquid air storing system must be optimized on a case-by-case basis, according to actual conditions. For extreme conditions, such as high peaks for short periods the other methods of peak shaving such as high pressure storage or liquid oxygen storage, without liquid air storage must be investigated and compared.

Concluding Remarks

In this paper we have tried to interpret and explain air separation plant performance and capital cost trends resulting from parametric process variations studies. They show that the purity of the product oxygen must be considered independently of the enriched air concentration required by the user. High pressure systems seem to be more energy efficient than low pressure systems if power recovery from pressurized product streams can be carried out efficiently. However, capital costs of high pressure process plants are higher than those of plants delivering product at essentially atmospheric pressure.

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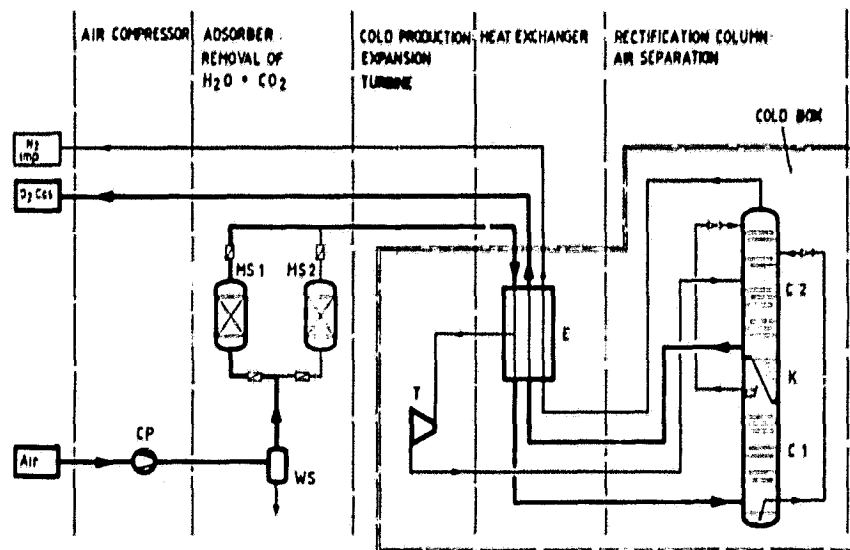
FIG. I   AIR SEPARATION PLANT

CONSTRUCTION

The columns are vertical cylindrical vessels containing sieve trays. The diagram below shows the basic constructional details of a sieve tray column.
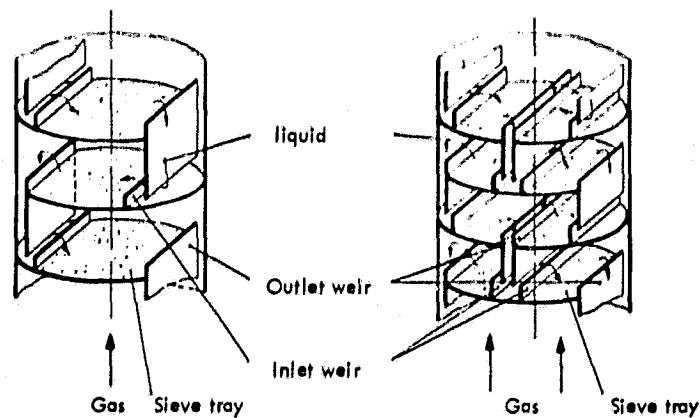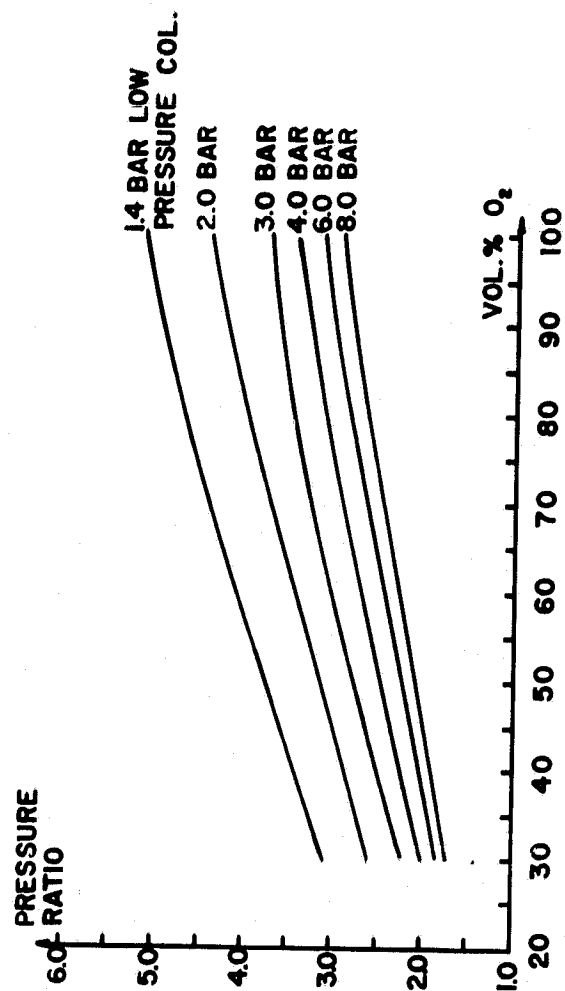

FIG. 2   RECTIFICATION COLUMN DETAILS

FIG. 3

PRESSURE RATIO

6.0
5.0
4.0
3.0
2.0
1.0

20 30 40 50 60 70 80 90 100

VOL. % O_2

1.4 BAR LOW PRESSURE COL.
2.0 BAR
3.0 BAR
4.0 BAR
6.0 BAR
8.0 BAR

PRODUCT OXYGEN CONCENTRATION

FIG. 3   PRESSURE RATIO: $\dfrac{\text{AIR}}{\text{PRODUCT-}O_2}$

OF AN AIR SEPARATION PLANT

FIG. 4

η_i

0.79
0.78
0.77
0.76
0.75
0.74
0.73
0.72
0.71
0.70

2   6   10   14   18   22   26

PRESSURE RATIO

FIG. 4   ISOTHERMAL EFFICIENCY OF
LARGE COMPRESSORS

Compressed
Air

Atm. Air (A) → CP → @ p_A → Air Separation Unit (ASU)

N_2 Fraction 1 (N1)

N_2 Fraction 2 (N2)

Product Oxygen (O)

Blend Air (AA)

Enriched Air (E)

Definitions:

p = pressure (atm) P = power

Power Factor*: $\frac{P}{A} = \ln(p_A) - \frac{N1}{A} \ln(p_{N1}) - \frac{N2}{A} \ln(p_{N2}) - \frac{O}{A} \ln(p_O)$

Air Factor: $\frac{A}{O}$ = (Inverse Recovery)

Product Factor: $\frac{O}{E}$

Specific Energy: $\frac{P}{E} = \frac{P}{A} \ast \frac{A}{O} \ast \frac{O}{E}$

* Negative terms indicate power recovery.

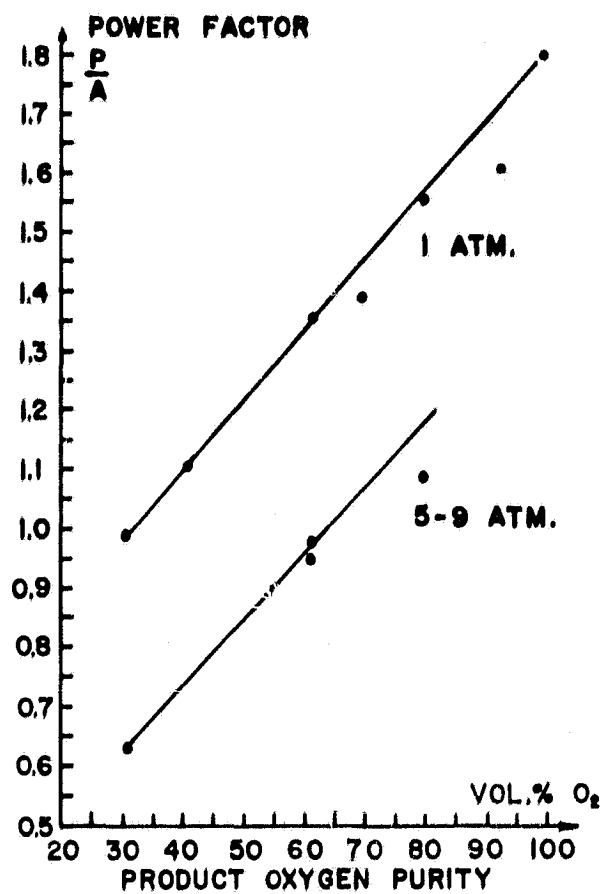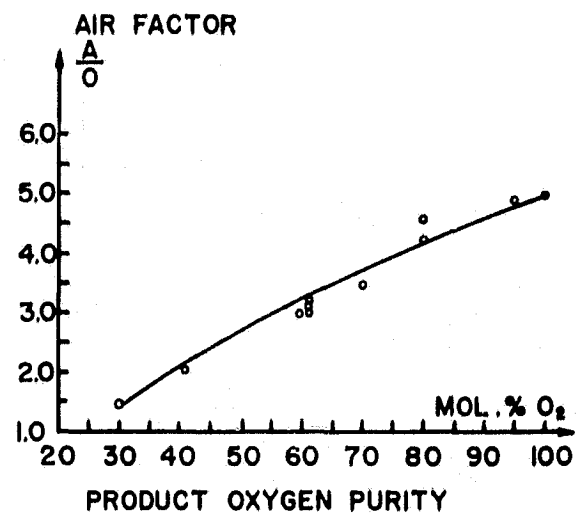FIG. 5 Definition of Specific Energy Consumption Terms

POWER FACTOR

1.8 $\frac{P}{A}$

1.7

1.6

1.5   1 ATM.

1.4

1.3

1.2

1.1   5-9 ATM.

1.0

0.9

0.8

0.7

0.6   VOL.% O_2

0.5
   20 30 40 50 60 70 80 90 100
   PRODUCT OXYGEN PURITY

FIG. 6   POWER FACTOR ASSUMING
          ISOTHERMAL COMPRESSION

AIR FACTOR
$\frac{A}{O}$

6.0

5.0

4.0

3.0

2.0   MOL.% O_2

1.0
   20 30 40 50 60 70 80 90 100
   PRODUCT OXYGEN PURITY

FIG. 7 AIR FACTOR

PRODUCT OXYGEN PURITY

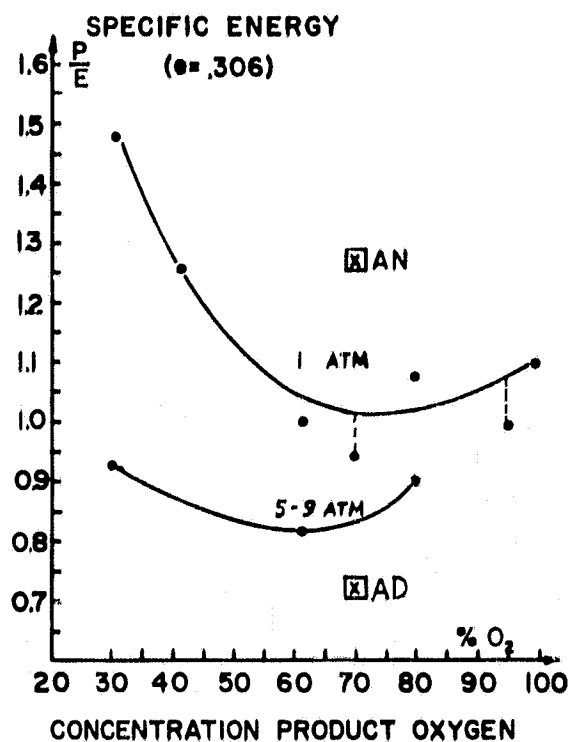FIG.8   PRODUCT FACTOR

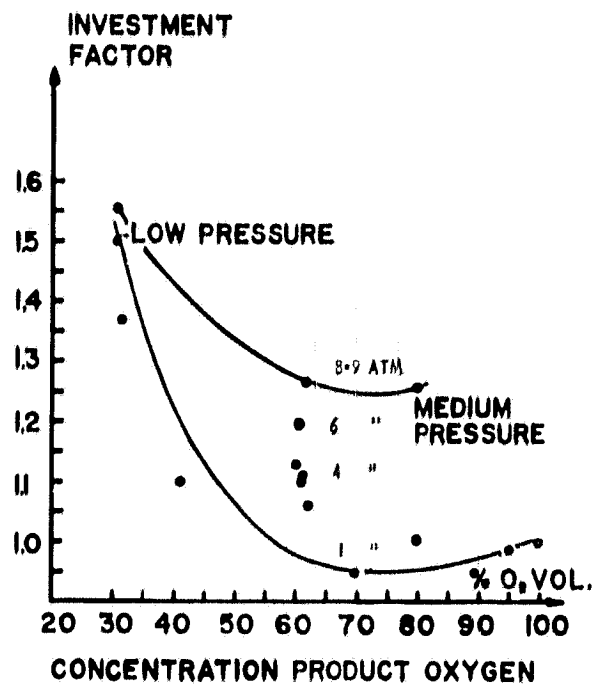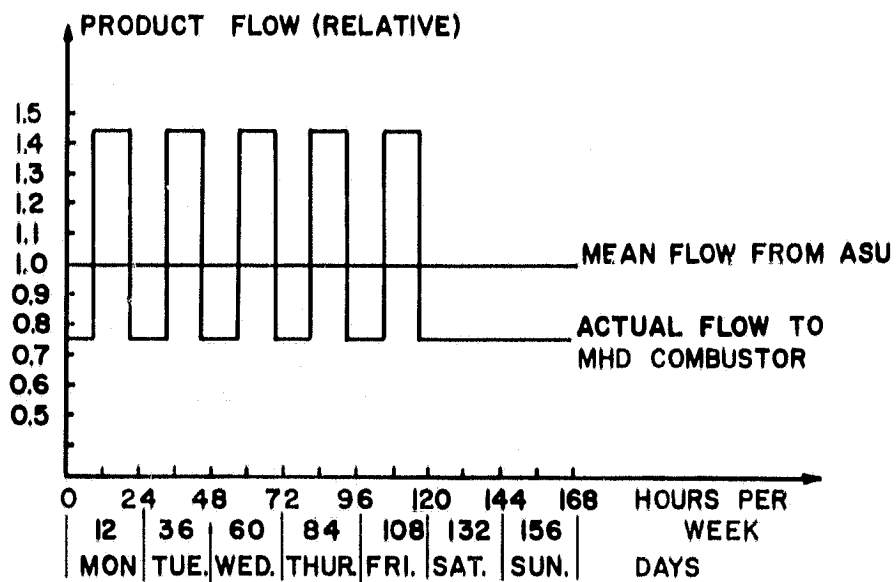CONCENTRATION PRODUCT OXYGEN

FIG.9   SPECIFIC ENERGY $\frac{P}{E}$

FIG.10 INVESTMENT COSTS

FIG. 11 ALTERNATING OXYGEN DEMAND AND PRODUCTION